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STEAM BOILERS, POWER-GENERATING FUEL, BURNERS, AND BOILER AUXILIARY EQUIPMENT

Transients in a Circulating Fluidized Bed Boiler

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Abstract—Transients in a circulating fluidized bed boiler firing biomass are considered. An attempt is made to describe transients with the use of concepts applied in the automatic control theory. The parameters calculated from an analysis of unsteady heat balance equations are compared with the experimental data obtained in the 12-MW boiler of the Chalmers University of Technology. It is demonstrated that these equations describe the transient modes of operation with good accuracy. Dependences for calculating the time constants of unsteady processes are obtained.

Keywords: fluidized bed, circulating fluidized bed, transients, biomass, boiler, circulation ratio, temperature
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Operation of circulating fluidized bed (CFB) boilers is described in detail in many individual publications, as well as in monographs, e.g. [1–3]. However, transients in such boilers are addressed only in a few works [4–6]. These processes need a more detailed study, in particular, for development of automatic closed-loop control systems. An attempt is made in this work to describe transients using the automatic closed-loop control analysis methods. The parameters calculated during an analysis of suitable unsteady equations are compared with the experimental data obtained in the 12-MW boiler of the Chalmers University of Technology in Sweden.

DESCRIPTION OF BOILER OPERATION

The 13.5-m-high boiler furnace is made of membrane-type waterwalls (Fig. 1). The waterwalls installed on the two opposite walls in the furnace lower part with a cross section of 1.4×1.4 m^2 and higher in the transport zone with a cross section of 1.4×1.7 m^2 are lined with refractory material. The cyclone with an inner diameter of 2 m is lined from inside with the same material. The heat-transfer surface in the transport zone has an area of 30 m^2. Particles from the cyclone enter into a lock composed of fluidized particles, from which they return into the furnace in bypass of an external cooler of particles, which was not used in these experiments (in firing biofuel). The boiler is fitted with a system for acquiring primary data that allows information to be received from several hundred measurement points. Sand with an average size of particles equal to 0.3 mm, which is used as bed inert material, is fluidized by primary air, and fluidizing air from the lock is also fed to the furnace. Secondary air is admitted into the furnace through nozzles placed at a height of 2.2 m from the air distribution grate.

The combustion products moved in the furnace with a velocity essentially higher than the terminal velocity of the majority of sand particles. This circumstance led to intense agitation of particles and strong circulation over the furnace–cyclone–lock–furnace loop. The experiments that had previously been carried out on that boiler showed the following.

(i) A fluidized bed with porosity $\varepsilon \approx 0.6$ exists in the furnace bottom part; its height $H_{f.b}$ was maintained by supplying sand and was equal to 0.7 m.

(ii) A zone of splashes locates above the bottom layer, the average concentration of particles in which is equal to around 100 kg/m^3; the height of this zone reaches 1.0–1.5 m from the air distribution grate.

(iii) A transport zone locates above the zone of splashes; the average density of particles in this zone varies in the range 1–20 kg/m^3 and depends on the primary air velocity.

During operation on biomass at the nominal load (8 MW) and at 80% of the nominal load, the temperature at the furnace outlet is 20–30°C higher than that in the bottom layer. This means that, although fuel is supplied on the bottom layer surface, it burns not only in the bottom layer (in which semicoke combustion predominantly takes place), but also in the zone of splashes, in which the majority of volatiles burn out, but part of the heat releasing during this process is transferred downward into the bottom layer with the particles returning into it. Depending on the size of fuel particles and agitation conditions, part of fuel can also burn in the transport zone. The material returning into the furnace from the cyclone has a temperature lower than the temperature at the furnace outlet, because part of heat is transferred through the cyclone refractory insulation to water-cooled tubes, and another part of heat is spent for heating the fluidizing

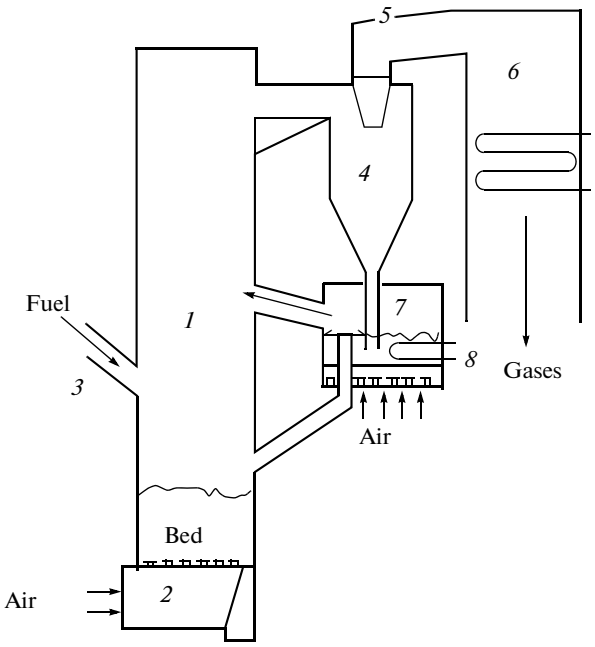

Fig. 1. Schematic design of the 12-MW CFB boiler of the Chalmers University of Technology. (*1*) Furnace, (*2*) air distribution chamber with a starting burner, (*3*) fuel feed, (*4*) cyclone, (*5*) outlet of gases to the convective part, (*6*) convective part, (*7*) fluidized lock, and (*8*) external heat exchanger.

air in the lock (the external heat exchanger was disconnected in the course of these experiments on biofuel).

Wood chips with the determining size (thickness) $\delta = 8-10$ mm were used as biofuel. They had the following elemental composition (in dry state, the content of moisture in working state W^r), %:

$$W^r = 37.2; \quad C^c = 51.1; \quad H^c = 6.3;$$
$$S^c = 0.03; \quad O^c = 41.4; \quad A^c = 0.8.$$

The wood chip combustion reaction rate constant found at laboratory conditions at a temperature of particles $T_p \geq 1073$ K is described by the formula $k = 288\exp(-8540/T_p)$ m^{-1} and at $T_p < 1073$ K by the formula $k = 1.9\exp(-2650/T_p)$ m^{-1}.

HEAT BALANCE OF THE FURNACE UNDER STEADY-STATE CONDITIONS

The wood chip combustion process can be subdivided into three consecutive stages: drying, release and combustion of volatiles, and afterburning of coke residue. Since fuel particles have approximately the same sizes exceeding the terminal sizes at the given gas velocity in the bottom layer, it can be considered that the chips are dried and the volatiles release from them in the bottom layer or at least on its surface. The volatiles burn out predominantly in the splash zone, in which they are mixed with the oxygen of air admitted

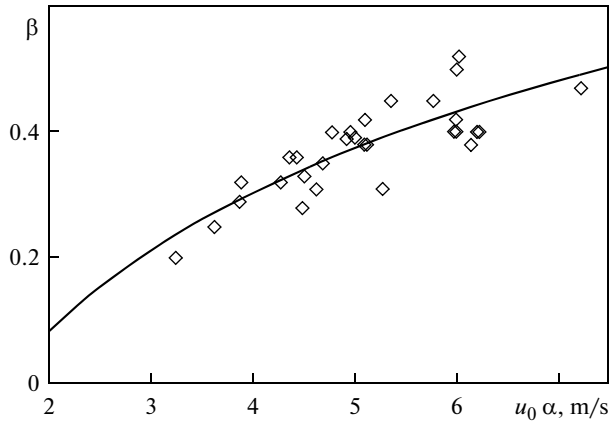

Fig. 2. Fraction of heat β from combustion of volatiles entering into the bottom layer vs. the $u_0\alpha$ complex.

through the bottom layer in the form of bubbles and also with the oxygen of air admitted with secondary blasting. Part of the heat releasing during the combustion of volatiles β is transferred from the splash zone into the bottom layer with the particles falling into it. The value of β was found from the heat balances of the furnace and bottom layer in the course of processing experimental data (Fig. 2).

The following flows of heat enter into the furnace under steady-state conditions:

(i) the heat releasing during the combustion of fuel;

(ii) the heat of primary and secondary air with the temperature $t_a \approx 70°C$;

(iii) the heat of recirculation gases having the temperature $t_r \approx 70°C$; and

(iv) the heat of air used for fluidization of particles in the lock and heated to the temperature t_l of the material returning into the furnace.

The heat releasing in the furnace is spent for increasing the temperature of combustion products to the temperature t_{tr}, transferred to the heat-absorbing surfaces, and spent for heating the recirculating material. The specific heat appearing in the equations is understood to mean the average specific heat either in the range from zero to the specified temperature or in the temperature range from t_l to t_{tr}. The heat balance equation looks as follows:

$$BQ_l^r + (G_1 + G_2)c_a t_a + G_r c_g t_r + G_l c_g t_l = F_{tr} w c_g t_{tr} + kF_{h.t}(t_{tr} - t) + G_{in} c_{in}(t_{tr} - t_l), \quad (1)$$

where B is the fuel flowrate, kg/s; Q_l^r is the fuel lower heating value in the working state, MJ/kg; G_1, G_2, G_l, and G_r are the flowrates of primary air, secondary air, air supplied to the fluidized lock, and recirculation gases, m^3/s; G_{in} is the flowrate of circulating material, kg/s; c_g and c_a are the per unit volume heat capacities

of gas and air, kJ/(m^3 K); c_{in} is the per unit mass heat capacity of the bed inert particles, kJ/(kg K); k is the heat-transfer coefficient in the furnace, W/(m^2 K); w is the velocity of gases in the furnace recalculated for the standard temperature and pressure, m/s; and $F_{h.t}$ is the heat-transfer surface area in the furnace, m^2.

The heat balance of the bottom layer is composed of the heat entering from coke residue combustion, the heat of a part of volatiles, and the heat of primary air, air from the lock, and recirculation gases. The entering heat is spent to evaporate moisture from the fuel and to heat the combustion products and recirculating particles:

$$B\frac{C_c}{100}Q_c + Q_l + \beta B\frac{V^r}{100}Q_v + G_r c_g t_r + G_l c_g t_l + G_l c_a t_a = B\frac{W^r}{100}r + c_g w_0 F_{f.b} t_{f.b} + G_{in} c_{in}(t_{f.b} - t_l), \quad (2)$$

where V^r is the yield of volatiles in the working state, %; C_c is the content of carbon in semicoke, %; r is the heat of vaporization, kJ/kg; and $t_{f.b}$ is the bed temperature.

The second term in the left-hand side of the equation describes the heat releasing as a result of semicoke combustion in the lock. The coefficient of heat transfer to the heat-transfer surfaces in the furnace $k_{h.t}$, W/(m^2 K) was calculated from the empirical formula $k_{h.t} = 85(\overline{\rho})^{0.3}$ obtained from the results of experiments on studying heat transfer in the furnaces of large CFB boilers (also in the boiler shown in Fig. 1) [8, 9] as a function of the average density $\overline{\rho}$, kg/m^3 of suspension in the transport zone, in which the heat absorption surfaces (waterwalls) are placed. In turn, the suspension density depends on the external circulation intensity of bed material in accordance with the following formula [7–10]:

$$\overline{\rho} = \frac{G_{in}}{F_{tr}(H_{tr} - H_{f.b})K_a(u - u_{trm})} \times \{\exp[K_a(H_{tr} - H_{f.b})] - 1\}, \quad (3)$$

where F_{tr} is the cross-section area of the furnace upper part, m^2; $H_{f.b}$ is the furnace and bottom layer height, m; $K_a = 0.23/(u - u_{trm})$ is the empirical coefficient describing the carry-over of material from the splash zone, 1/m; u is the gas velocity at the furnace outlet, m/s; and u_{trm} is the terminal velocity of a free particle, m/s.

For boiler operation under steady-state conditions at different loads and with the parameters indicated above, Eqs. (1)–(3) give expressions for the material recirculation intensity $G_{in} = 0.03u_0^{3.3}$ as a function of the fluidization velocity u_0 in the bottom layer, as well as the fraction of heat absorbed by the bottom layer during the combustion of volatiles $\beta = 0.31\ln(u_0\alpha) - 0.132$ [11], where α is the primary air excess factor.

As a semicoke particle burns out, its volume decreases by several hundred times before its size decreases to approximately 1 mm, which corresponds to its free hovering at the given gas velocity. Thus, we can assume that wood semicoke burns out mainly in the bottom layer and in the splash zone. Hence, it becomes possible to simplify the model and to assume that fuel particles do not burn in the transport zone, which is confirmed by the fact that the semicoke concentration remains constant along the furnace height.

An unexpected result was obtained from assessments aimed at determining the possibility of coke to burn in a fluidized-bed lock. Namely, with the particle circulation velocity related to the furnace transport zone cross section equal to around 10 kg/(m^2 s), carbon content equal to 0.3%, and air flowrate through the lock equal to 0.15 m^3/s, the corresponding flow air excess factor is much less than unity (0.3 at the nominal load). The actual air excess factor calculated from the composition of combustion products depends on the amount of semicoke that actually burns out in the lock. Its value is determined by the bed height, temperature, specific surface area, and reaction capacity of semicoke particles. In the considered case, the temperature in the fluidized-bed lock (750°C) was sufficient for semicoke particles to burn in it. The size these particles had at the furnace outlet δ_c found from screening was equal to 0.6 mm. With the measured concentration of semicoke particles z_c, their specific surface area in the lock can be calculated from the formula

$$S_l = 6\rho_{in}(1 - \varepsilon)z_l/(\rho_c\delta_c).$$

Knowing the semicoke reaction capacity and using the information presented above, we can easily calculate the carbon dioxide concentration $\overline{C}_{CO_2}$ in the gas leaving the lock (its flowrate is equal to G_l) and, hence, the amount of semicoke that burns out in the lock

$$m_l = G_l\rho_c\frac{12}{44}\frac{C_{CO_2}}{100}.$$

Taking the heating value $Q = 32.6$ MJ/kg, we can calculate the amount of heat releasing in the lock during the combustion of semicoke particles in it:

$$Q_l = Qm_l.$$

DIFFERENTIAL EQUATIONS OF A TRANSIENT

For describing transients in a CFB boiler, we should write heat and mass balance equations. To do so, we take the following assumptions.

(i) The changes of gas concentrations caused by a change in the concentrations $z_{f.b}$ of combustibles in the bottom layer volume $V_{f.b}$ take place almost instantaneously as compared with the rate of change in the bed temperature, due to which the concentrations of gas-

eous components at each moment of time can be calculated from the equations of steady combustion mode (a quasi steady-state approach).

(ii) Owing to intense agitation of particles in the bottom layer, the combustibles are uniformly distributed in its volume, and this layer is isothermal in nature.

(iii) The volatiles burn out in the bottom layer and in the splash zone at a rate incommensurably faster than the rate at which they release from fuel particles. The fraction β of the volatiles combustion heat received by the bottom layer is presented in Fig. 2.

(iv) The combustion process in the furnace continues until the formation of CO_2, because in none of the studied modes was any noticeable amount of CO observed at the furnace outlet.

(v) The formation of CO during the combustion in the fluidized lock at low temperature is not taken into account, although at such temperature CO and not CO_2 is the primary product of the reaction $C + O_2$, which, as experiments show, burns out in the bed itself.

(vi) Owing to intense internal recirculation, the temperature t_{tr} in the transport zone is constant along the furnace height, but it differs from the bottom bed temperature $t_{f.b}$.

(vii) Experiments have shown that the time of transients connected with the change of the ratio between the mass of inert material in the bottom layer and its mass in the furnace above the bottom layer is small as compared with the time of temperature variation; therefore, this ratio of the masses of inert material was calculated from expressions obtained for steady-state modes (a quasi steady-state approach).

The unsteady equations for material balances of semicoke and fuel particles having a considerable content of volatiles can be written as follows:

$$M_{f.b}\frac{dz_{f.b}}{d\tau} = B\frac{C_l}{100} - V_{f.b}j_{f.b}\frac{6\rho_{in}(1-\varepsilon)}{\rho_l\delta_{f.b}}z_{f.b} - G_{in}(f_{tr}z_{f.b} - z_l); \quad (4)$$

$$M_l\frac{dz_l}{d\tau} = G_{in}(f_{f.b}z_{f.b} - z_l) - m_l; \quad (5)$$

$$M_{f.b}\frac{dz_v}{d\tau} = B\frac{V^r}{100} - V_{f.b}j_v\frac{6\rho_{in}(1-\varepsilon)}{\rho_l\delta_0}z_v, \quad (6)$$

where $M_{f.b}$ and M_l are the masses of particles in the bottom layer and in the lock, kg; τ is time, s; and j are the flowrates of carbon from bottom layer unit surface and volatiles, kg/(cm m^2).

The last term in Eq. (4) reflects the external circulation of semicoke particles that are carried away from the bottom layer with the flowrate G_{in} and finally return into the furnace from the lock with fluidized bed, in which their concentration is equal to z_l. The concentration $f_{tr}z_{f.b}$ of semicoke in the flow G_{in} of material carried away from the bottom layer differs from the concentration $z_{f.b}$ in the layer itself, because large particles are not carried away from it. Taking the average size of semicoke particles in the bottom layer (found from screening) equal to $\delta = 3$ mm and using a linear approximation of the particle diameter corresponding to the particle terminal velocity [9], we can obtain an expression for relative concentration of particles carried away from the bed in the following form:

$$f_{tr} = 1 - 0.53(0.41 + 0.13u)^{-0.27}. \quad (7)$$

The mass of inert material in the bottom layer $M_{f.b}$ can be calculated as the difference between the total mass in the furnace M_Σ, which is determined with sufficient accuracy by the pressure difference and mass in the transport zone M_{tr}:

$$M_{f.b} = M_\Sigma - M_{tr}, \quad (8)$$

where $M_{f.b} = H_{f.b}F_{f.b}\bar{\rho}$. The average density $\bar{\rho}$ is calculated from (3).

The flow of carbon from unit surface of a semicoke particle reacting in the bottom layer is given by

$$j = 0.375\left(\frac{1}{k} + \frac{1}{\alpha_d}\right)^{-1}\bar{C}_{O_2}, \quad (9)$$

where the diffusion mass transfer coefficient of the reacting semicoke particle $\alpha_d = \mathrm{Sh}\,D/\delta = 0.375$ is a stoichiometric coefficient, kg C/kg O_2; Sh is the Sherwood number; and $\bar{C}_{O_2}$ is the mass concentration of oxygen. The coefficient of oxygen diffusion in nitrogen is given by the following formula [12]:

$$D = 0.16\times10^{-4}\left(\frac{t_{f.b}}{273}\right)^{1.9}. \quad (10)$$

The Sherwood number is calculated from the following expression [13]:

$$\mathrm{Sh} = \frac{\delta}{\delta_{in}}\left\{2.78\,\mathrm{Re}_k^{0.33}\left(\frac{\delta_{in}}{\delta}\right)^{0.67} + 0.45\,\mathrm{Re}_k^{0.56}\left[1 - \left(\frac{\delta_{in}}{\delta}\right)^{0.67}\right]\right\}\mathrm{Sc}^{0.33}, \quad (11)$$

where Sc is the Schmidt number, and $\mathrm{Re}_k = \mathrm{Ar}\big/\left(1400 + 5.22\sqrt{\mathrm{Ar}}\right)$ (Ar is the Archimedean number) is the Reynolds number for the onset of fluidization [14].

The average active concentration of oxygen in the bottom layer is calculated in accordance with the following formula [15]:

$$\bar{C}_{O_2} = \frac{c_0u_0}{\alpha_r S_{f.b}H_{f.b}}\left[1 - \exp\left(\frac{\alpha_r S_{f.b}H_{f.b}}{u_0}\right)\right], \quad (12)$$

where $S_c = \dfrac{6\rho_{in}(1-\varepsilon)z_{f.b}}{\rho_c\delta}$ is the specific surface of reacting semicoke in the bottom layer, which depends on the mass concentration $z_{f.b}$ of semicoke; c_0 is the

concentration of oxygen in the air admitted to under the bed, kg/m^3; and $\alpha_r = 1\Big/\left(\frac{1}{k}+\frac{1}{\alpha_d}\right)$ is the reaction gas transfer coefficient.

Equation (5) describes the change in the mass of particles in the lock, and Eq. (6) describes the change in the mass of particles in the bottom layer containing volatiles. Equation (6) can be excluded from the system of equations without any noticeable loss of accuracy, because the time taken for the volatiles to release and burn out is around 1 min, and the time of transient is 10–20 min. In the general case, the flow of volatiles j_v releasing from unit surface of fuel particles is calculated in accordance with [16] from the following expression:

$$j_v = \frac{V^r}{100\,\tau_v}\rho_r, \tag{13}$$

where ρ_r is the density of fuel particles, kg/m^3, and τ_v is the time taken for volatiles to release from the wood, which was determined experimentally [11].

The equations for unsteady modes of heat balances describe variations of temperature in the bottom layer, at the furnace outlet, and in the lock with fluidized bed:

$$\begin{aligned} M_{f.b}c_{in}\frac{dt_{f.b}}{d\tau} &= V_{f.b}\frac{6\rho_{in}(1-\varepsilon)}{\rho_c\delta}jz_{f.b}Q_c \\ &+ \beta V_{f.b}\frac{6\rho_{in}(1-\varepsilon)}{\rho_r\delta_0}j_v z_v Q_v \\ &+ G_l c_a t_a + G_r c_g t_r - B\frac{W^r}{100}r - w_0 c_g F_{f.b} t_{f.b} \\ &- G_{in}c_{in}(t_{f.b}-t_l) + G_l c_g t_l; \end{aligned} \tag{14}$$

$$\begin{aligned} V_{tr}(c_g + c_{in}\bar{\rho})\frac{dt_{tr}}{d\tau} &= (1-\beta)BQ_v V^r/100 + G_2 c_a t_a \\ &+ w_0 c_g F_{f.b} t_{f.b} - w c_g F_{tr} t_{tr} \\ &- G_{in}c_{in}(t_{tr}-t_{f.b}) - k_{h.t}F_{h.t}(t_{tr}-t); \end{aligned} \tag{15}$$

$$\begin{aligned} M_l c_{in}\frac{dt_l}{d\tau} &= G_{in}c_{in}(t_{tr}-t_l) \\ &+ G_l c_a t_a - G_l c_g t_l - k_l F_l (t_l - t) + Q_l, \end{aligned} \tag{16}$$

where W^r is the moisture content in fuel in its working state, %; w_0 is the initial velocity, m/s; t is the coolant temperature (water heated in the boiler), °C; $F_{f.b}$ is the cross-section area of the furnace lower part, m^2; $F_{h.t}$ is the heat-transfer surface area in the furnace, m^2; F_l is the heat-transfer surface area in the intermediate chamber, m^2; k_l is the heat-transfer coefficient in the lock, W/(m^2 K); Q_c and Q_v are the lower heating values of semicoke and volatiles, MJ/kg; z_v is the mass fraction of combustibles in the bed containing volatiles; δ_0 is the initial size of fuel particles, m; V_{tr} is the volume of the transport zone, m^3; and ρ_{in} is the density of inert particles, kg/m^3.

CALCULATION OF TIME CONSTANTS

In the automatic control theory, the variation rate of parameters during transients is characterized by time constants, which appear as coefficients of derivatives in the differential equations of control elements written in standard form and having the dimension of time. These constants are determined by calculation or experimentally by taking transient responses [17].

An analysis of formulas (4)–(16) yields expressions for the time constant main components that take into account the following:

—heating of material in the bottom layer

$$T_{f.b} = \frac{M_{f.b}\,c_{in}}{G_l c_a + G_{in}c_{in} - G_{in}c_{in}\dfrac{G_l c_a + G_{in}c_{in}}{G_{in}c_{in}+G_l c_a + k_{h.t}F_{h.t}}\dfrac{G_{in}c_{in}}{G_{in}c_{in}+k_l F_l}}; \tag{17}$$

—heating of material above the bottom layer

$$T_{tr} = \frac{V_{tr}(\bar{\rho}c_{in}+c_g)}{k_{h.t}F_{h.t} + G_l c_g + G_{in}c_{in}\dfrac{k_l F_l}{G_{in}c_{in}+k_l F_l}}; \tag{18}$$

—cooling of material in the lock

$$T_l = \frac{M_l c_{in}}{k_{h.t}F_{h.t} + G_{in}c_{in}\dfrac{kF_p + G_l c_g}{k_c F_c + G_l c_g + k_{h.t}F_{h.t}}}; \tag{19}$$

—semicoke combustion time constant, which is in fact the combustion time of its particle

$$T_c = \frac{\delta\,\rho_c}{6j}, \tag{20}$$

where ρ_c is the semicoke density, kg/m^3; and

—combustion of volatiles

$$T_v = \frac{\delta_0\rho_p}{6j_v}. \tag{21}$$

The time constant of the resulting transient is equal to the sum of components. However, the influence of two latter ones is sometimes not obvious, because the semicoke and volatile combustion flows depend both

Limiting values of the time constant components

Boiler zone	Time constant	
	$G_{in} = 0$	$G_{in} \to \infty$
Bottom layer	$\frac{M_{f.b}c_{in}}{c_a G_l}$	$\frac{M_{f.b}c_{in}}{c_a G_l + k_p F_p}$
Transport zone	$\frac{V_{tr}(\bar{\rho}c_{in} + c_a)}{G_l c_a + k_p F_p}$	$\frac{V_{tr}(\bar{\rho}c_{in} + c_a)}{G_l c_a + k_p F_p + k_c F_c}$
Lock	Does not have a physical sense	$\frac{M_c c_{in}}{G_l c_a + k_p F_p + k_c F_c}$

on the temperature and on the change of oxygen concentration during the transient. The table gives the limiting values of the time constant components obtained from expressions (17)–(19) at $G_{in} = 0$ (a stationary fluidized bed) and at $G_{in} \to \infty$ (a CFB with very intense external circulation of particles).

It can be seen from an analysis of expressions (17)–(19) that the time constants depend directly on the circulation rate: they decrease as G_{in} increases (which is determined by the primary air velocity). Moreover, the inert particle flow density (i.e., the mass of inert particles in the transport zone) depends on the value of G_{in} in the freeboard, which in turn determines the heat transfer in the furnace. Hence, the primary air velocity is one of the key parameters causing the CFB boiler inertia properties.

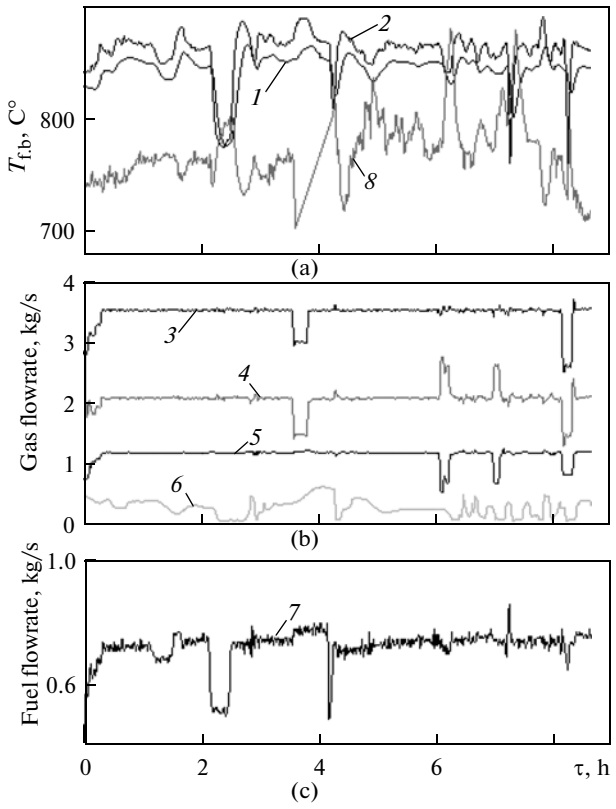

Fig. 3. Variation of furnace parameters an hour after applying disturbances during operation at the nominal load. (*1*) Bottom layer temperature, (*2*) temperature at the furnace outlet, (*3*) total air flowrate, (*4*) primary air flowrate; (*5*) secondary air flowrate, (*6*) flowrate of recirculation gases, (*7*) fuel flowrate, and (*8*) temperature in the lock with fluidized bed.

COMPARISON BETWEEN THE RESULTS OF CALCULATIONS AND EXPERIMENTS IN A CFB BOILER

Figure 3 shows variations of the bottom layer temperature, temperature at the furnace outlet, and in the lock in response to disturbances applied in the furnace during its operation at the nominal load by reducing the fuel feed at constant air flowrate, by changing primary air flowrate at constant secondary air and fuel flowrates, by increasing primary and decreasing secondary air flowrate while keeping constant total air and fuel flowrates, and by decreasing the total air flowrate while keeping constant fuel flowrate and the ratio of primary to secondary air flowrates. A decrease of fuel flowrate (see Fig. 3) causes a gradual drop of temperature in the bottom layer and at the furnace outlet. The time required for reaching the new steady-state operating conditions in the transport zone is significantly shorter than that in the bottom layer because the major part of heat releases in the splash zone during the combustion of volatiles in it. Part of the heat released during the combustion of fuel is transferred by gas and particles into the transport zone, and another part of this heat is transferred into the bottom layer. The bottom layer is heated more slowly because the mass of inert material in it is larger than it is in the transport zone. The regularities pertinent to variation of temperature in the lock with fluidized bed are more intricate in nature. After a decrease of fuel flowrate, the temperature in the lock first becomes somewhat lower due to a drop of the circulating material temperature at the furnace outlet, after which the temperature in it increases because the fuel flowrate remains almost constant, whereas the heat release increases due to an increase of semicoke content in the bottom layer and, hence, in the circulating material that enters into the lock.

An increase of fuel flowrate at a constant air flowrate exceeding the amount theoretically necessary for complete combustion of fuel leads, naturally, to a growth of temperature in the entire furnace volume. A decrease of primary air flowrate at a constant secondary air flowrate affects the change of temperature in the same way as a decrease of fuel flowrate, because in this case we have a smaller amount of combustion products to be heated.

Figure 4 shows the dependences of time constant on the velocity at the furnace outlet calculated from

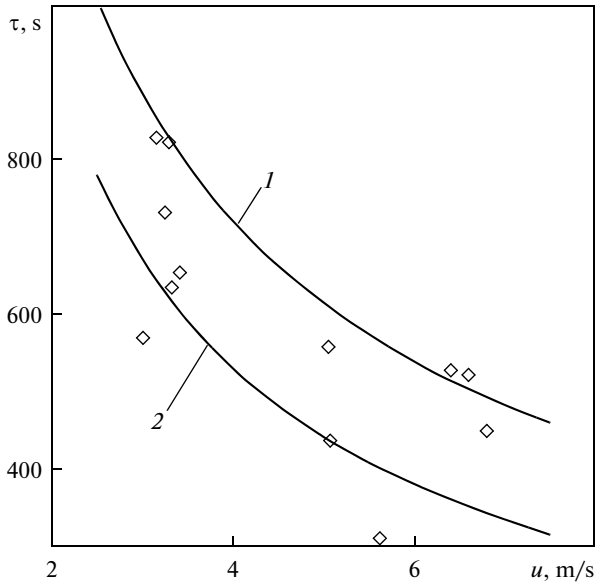

Fig. 4. Comparison between the calculated and experimental values of time constant as a function of velocity at the furnace outlet. The curves are for the calculated values, and the points are for the experimental values. (*1*) The sum of all time constants and (*2*) the time constant determined without taking into account the combustion time of semicoke and volatiles.

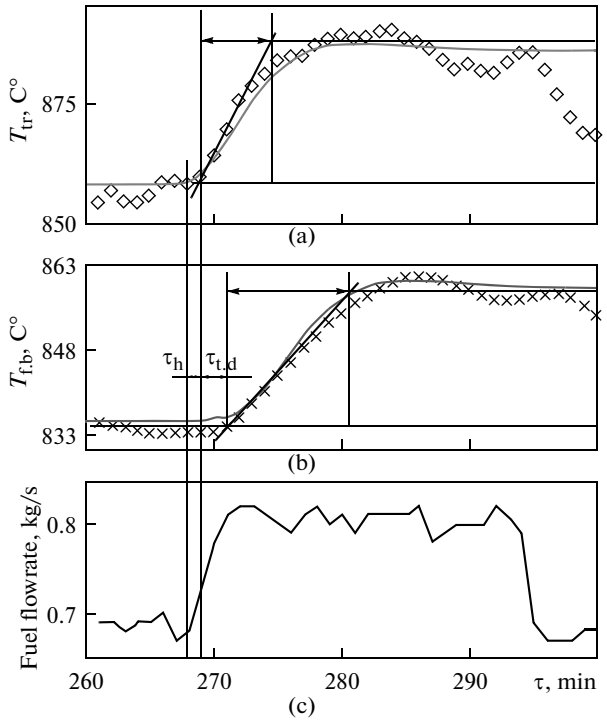

Fig. 5. Comparison between the experimental (points) and calculated (curves) temperatures at the furnace outlet (a) and in the bottom layer (b) after application of a disturbance by changing the fuel flowrate (c).

Eqs. (17)–(21) and (17)–(19) without taking into account the components represented by Eqs. (20) and (21). As the gas velocity is increased from 3 m/s (which corresponds to boiler operation at 60% of its nominal load) to 6.5 m/s (operation at the nominal boiler load), the system inertia drops by almost a factor of 2. Figure 4 also shows the time constants calculated experimentally from the temperature variation rate. Almost all experimental points lie between the two calculated curves, from which it follows that the model can be used for analyzing real processes.

Figure 5 shows variations of the bottom layer temperature and the temperature at the furnace outlet in response to changing the fuel flowrate from 0.7 to 0.8 kg/s in comparison with the transient calculated from the obtained values of time constants. We see that the upper curve (variation of temperature in the freeboard) can be approximated by a first-order inertial section and T_{tr} is numerically equal to the plant time constant. The medium curve represents a complex plant with an S-shaped curve that is replaced by a first-order model with the time constant $T_{f.b}$ and plant delay time $\tau_0 = \tau_{p.d} + \tau_{t.d}$, where $\tau_{p.d}$ is the pure (transport) delay time and $\tau_{t.d}$ is the transient delay time.

The calculated dependences are in satisfactory agreement with the experimental data.

REFERENCES

1. *Circulating Fluidized Beds*, Ed. by J. R. Grace, A. A. Avidan, and T. M. Knowlton (Blackie Academic and Professional, 1997).
2. P. Basu and S. A. Frazer, *Circulating Fluidized Bed Boilers, Design and Operation* (Butterworth-Heinemann, Quebec, 1991).
3. A. P. Baskakov, V. V. Matsnev, and I. V. Raspopov, *Fluidized Bed Boilers and Furnaces* (Energoatomizdat, Moscow, 1995) [in Russian].
4. C. K. Park and P. Basu, "Operation of an existing pulverized coal fired boiler as a circulating fluidized bed boiler," Fluidized Bed Combustion, No. 1, 77–83 (1987).
5. C. K. Park and P. Basu, "A model for prediction of transient response to the change of fuel feed rate to a circulating fluidized bed boiler furnace," Chem. Eng. Sci., No. 52, 3499–3509 (1997).
6. T. Hyppanen, A. Kettunen, Y. Y. Lee, and J. Riiali, "Dynamic modeling for simulation and control of a circulating fluidized-bed combustor," Fluidized Bed Combustion, No. 2, 1121–1127 (1993).
7. F. Johnsson and B. Leckner, "Vertical distribution of solids in a CFB furnace," in *Proceedings of the 13th International Conference on FBC, New York, American Society of Mechanical Engineers, Orlando, ASME*, pp. 266–273.
8. A. P. Baskakov, B. Leckner, and K. Breitkholts, "Complex heat transfer in CFB furnaces," in *Proceedings of*

the 2nd Russian National Conference on Heat Transfer, MEI, Moscow, 1998, Vol. 5, pp. 151–154.

9. M. Komatina, V. Manovich, and M. Paprika, "Heat transfer between a coal particle and fluidized bed during heating and devolatilization. Circulating fluidized bed boilers," in *Proceedings of the 9th International Conference on Circulating Fluidized Beds, Germany, Hamburg Technical University, May 13–16, 2008.*
10. Y. Teplitskii, V. Kovenskii, and V. Borodulya, "Numerical modeling of solid mixing in a CFB. Circulating fluidized bed boilers," in *Proceedings of the 9th International Conference on Circulating Fluidized Beds, Germany, Hamburg Technical University, May 13–16, 2008.*
11. V. A. Munts, E. Yu. Pavlyuk, M. V. Kryukov, and A. V. Savkin, "Combustion of volatiles during the firing of solid fuels in a fluidized bed," Elektr. Stn., No. 3, 16–22 (2012).
12. V. I. Babii and Yu. F. Kuvaev, *Combustion of Pulverized Coal and Calculation of a Pulverized-Coal Flame* (Energoatomizdat, Moscow, 1986) [in Russian].
13. V. A. Borodulya, G. I. Pal'chenok, and G. G. Vasil'ev, "Problems of heat and mass transfer in modern devices for firing and gasification of solid fuels," in *Proceedings of the International School–Seminar, Minsk, 1988*, Part 2, pp. 3–23.
14. O. M. Todes and O. B. Tsitovich, *Fluidized Bed Apparatuses* (Khimiya, Moscow, 1981) [in Russian].
15. V. A. Munts, A. P. Baskakov, and A. A. Ashikhmin, "Calculation of solid fuel devolatilization in a fluidized bed," Inzh. Fiz. Zh., No. 54, 432–438 (1988).
16. *Fundamentals of a Practical Combustion Theory*, Ed. by V. V. Pomerantsev (Energiya, Leningrad, 1973) [in Russian].
17. E. P. Stefani, *Principles of Calculating the Tuning Parameters of Devices for Control of Thermal Power Processes* (Energiya, Moscow, 1972) [in Russian].

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